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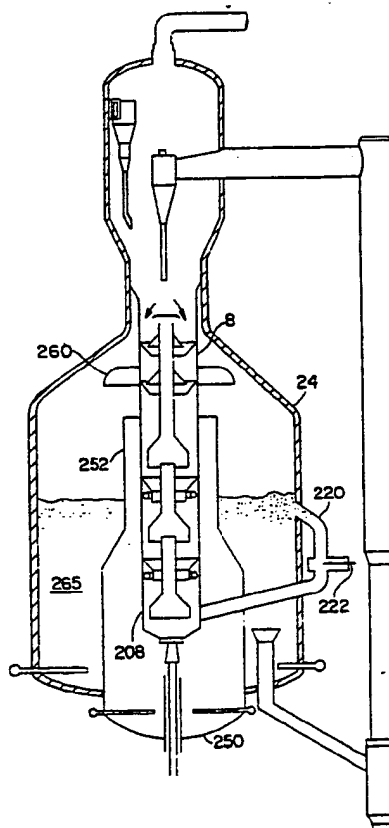
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(54) Title: A PROCESS FOR STRIPPING AND REGENERATING FLUIDIZED CATALYTIC CRACKING CATALYST**(57) Abstract**

A process is disclosed for achieving hot catalyst stripping of spent FCC catalyst in a bubbling bed regenerator (24) having a stripper (8) mounted over the regenerator and stripped catalyst standpipe (208) within the regenerator. Hot catalyst stripping is achieved by indirect heat exchange from the hot regenerated catalyst in the regenerator to the spent catalyst in the stripper. The indirect heat exchange is conveniently effected by immersing the hot stripper in the bubbling bed of the regenerator or in a coke combustor (250) associated with the regenerator, or by use of heat pipes containing a working fluid which is vaporized in the regenerator and condensed in the hot stripper. Regenerated catalyst may also be recycled (220) from the regenerator to the hot stripper to effect direct contact heating of the spent catalyst.



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A PROCESS FOR STRIPPING AND REGENERATING
FLUIDIZED CATALYTIC CRACKING CATALYST

The invention relates to a process and apparatus for stripping and regenerating fluidized catalytic cracking catalyst.

In the fluidized catalytic cracking (FCC) process, catalyst circulates between a cracking reactor and a catalyst regenerator. In the reactor, hydrocarbon feed contacts a source of hot, regenerated catalyst, which vaporizes and cracks the feed at a temperature of 425-600°C, usually 460-560°C. The cracking reaction deposits carbonaceous hydrocarbons or coke on the catalyst, thereby deactivating the catalyst. The cracked products are separated from the coked catalyst, which is then stripped of volatiles, usually with steam, in a catalyst stripper. The stripped catalyst is then passed to the catalyst regenerator, where coke is burned from the catalyst with oxygen containing gas, usually air. Decoking restores catalyst activity and simultaneously heats the catalyst to, for example, 500-900°C, usually 600-750°C. This heated catalyst is recycled to the cracking reactor to crack more fresh feed. Flue gas formed by burning coke in the regenerator may be treated for removal of particulates and for conversion of carbon monoxide, after which the flue gas is normally discharged into the atmosphere.

Catalytic cracking has undergone progressive development since the 1940's. The trend of development of the fluid catalytic cracking (FCC) process has been to all riser cracking and the use of zeolite catalysts. A good overview of the importance of the FCC process, and its continuous advancement, is provided in "Fluid Catalytic Cracking Report", by Amos A. Avidan, Michael Edwards and Hartley Owen, published in the January 8, 1990 edition of the Oil & Gas Journal. One modern,

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compact FCC design is the Kellogg Ultra Orthoflow converter, Model F, which is shown in Figure 1 of the accompanying drawings and also shown as Figure 17 of the January 8, 1990 Oil & Gas Journal article. Although this unit works well in practice, the stripping of spent catalyst is never as complete as desired by the refiner. In addition, FCC units are being pushed to accept poorer feeds, particularly feeds containing large amounts of resid. These growing demands placed on FCC units have exacerbated four existing problem areas in the regenerator, namely problems with sulfur, steam, temperature and NO_x . These problems will each be reviewed in more detail below.

SULFUR

Much of the sulfur in the feed ends up as SO_x in the regenerator flue gas. Higher sulfur levels in the feed, combined with a more complete regeneration of the catalyst in the regenerator increases the amount of SO_x in the regenerator flue gas. Some attempts have been made to minimize the amount of SO_x discharged to the atmosphere through the flue gas by including catalyst additives or agents to react with the SO_x in the flue gas. These agents pass with the regenerated catalyst back to the FCC reactor where the reducing atmosphere releases the sulfur compounds as H_2S . Suitable agents are described in U.S. Patent Nos. 4,071,436 and 3,834,031. Use of cerium oxide agent for this purpose is shown in U.S. Patent No. 4,001,375.

Unfortunately, the conditions in most FCC regenerators are not the best for SO_x adsorption. The high temperatures in modern FCC regenerators (up to 870°C) impair SO_x adsorption. One way to minimize SO_x in flue gas is to pass catalyst from the FCC reactor to a long residence time steam stripper, as disclosed in U.S. Patent No. 4,481,103. This process preferably steam strips spent catalyst at $500\text{--}550^\circ\text{C}$, which is

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beneficial but not sufficient to remove some undesirable sulfur- or hydrogen-containing components.

STEAM

Steam is always present in FCC regenerators although it is known to cause catalyst deactivation. Steam is not intentionally added, but is invariably present, usually as adsorbed or entrained steam from steam stripping or catalyst or as water of combustion formed in the regenerator.

Poor stripping leads to an increased level of steam in the regenerator, firstly from the adsorbed or entrained steam and secondly from hydrocarbons left on the catalyst due to poor catalyst stripping. Catalyst passing from an FCC stripper to an FCC regenerator contains hydrogen-containing components, such as coke or unstripped hydrocarbons adhering thereto. This hydrogen burns in the regenerator to form water and cause hydrothermal degradation.

Steaming of catalyst becomes more of a problem as regenerators get hotter. Higher temperatures greatly accelerate the deactivating effects of steam.

TEMPERATURE

Regenerators are operating at higher and higher temperatures. This is because most FCC units are heat balanced, that is, the endothermic heat of the cracking reaction is supplied by burning the coke deposited on the catalyst. With heavier feeds, more coke is deposited on the catalyst than is needed for the cracking reaction. The regenerator gets hotter, and the extra heat is rejected as high temperature flue gas. Many refiners severely limit the amount of resid or similar high CCR feeds to that amount which can be tolerated by the unit. High temperatures are a problem for the metallurgy of many units, but more importantly, are a problem for the catalyst. In the regenerator, the burning of coke and unstripped hydrocarbons leads to much higher surface temperatures on the catalyst

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than the measured dense bed or dilute phase temperature. This is discussed by Occelli et al in "Dual-Function Cracking Catalyst Mixtures", Ch. 12, Fluid Catalytic Cracking, ACS Symposium Series 375, American Chemical Society, Washington, D.C., 1988.

Some regenerator temperature control is possible by adjusting the CO/CO₂ ratio produced in the regenerator. Burning coke partially to CO produces less heat than complete combustion to CO₂. However, in some cases, this control is insufficient, and also leads to increased CO emissions, which can be a problem unless a CO boiler is present.

U.S. Patent No. 4,353,812, discloses cooling catalyst from a regenerator by passing it through the shell side of a heat-exchanger with a cooling medium through the tube side. This and similar approaches remove heat from the regenerator, but will not prevent poorly, or even well, stripped catalyst from experiencing very high surface or localized temperatures in the regenerator.

The prior art also uses dense or dilute phase regenerated fluid catalyst heat removal zones or heat-exchangers that are remote from, and external to, the regenerator vessel to cool hot regenerated catalyst for return to the regenerator. Examples of such processes are found in U.S. Patent Nos. 2,970,117, 2,873,175, 2,862,798, 2,596,748, 2,515,156, 2,492,948 and 2,506,123.

NO_x

Burning of nitrogenous compounds in FCC regenerators has long led to creation of minor amounts of NO_x, some of which were emitted with the regenerator flue gas. Usually these emissions were not much of a problem because of relatively low temperature, a relatively reducing atmosphere from partial combustion of CO and the absence of catalytic metals like Pt in the regenerator which increase NO_x production.

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Many FCC units now operate at higher temperatures, with a more oxidizing atmosphere, and use CO combustion promoters such as Pt. These changes in regenerator operation reduce CO emissions, but usually increase nitrogen oxides (NO_x) in the regenerator flue gas. It is difficult in a catalyst regenerator to completely burn coke and CO in the regenerator without increasing the NO_x content of the regenerator flue gas, so NO_x emissions are now frequently a problem. These problems are more severe in bubbling bed regenerators, because of relatively poor catalyst circulation (large stagnant regions in the dense bed) and the presence of large bubbles of regeneration gas which leads to localized high concentrations of oxygen, which increases NO_x emissions.

Recent catalyst patents include U.S. 4,300,997 and 4,350,615, both directed to the use of Pd-Ru CO-combustion promoter. The bi-metallic CO combustion promoter is reported to do an adequate job of converting CO to CO_2 , while minimizing the formation of NO_x .

U.S. 4,199,435 suggests steam treating conventional metallic CO combustion promoter to decrease NO_x formation without impairing too much the CO combustion activity of the promoter.

Process modifications are suggested in U.S. 4,413,573 and U.S. 4,325,833 directed to two-and three-stage FCC regenerators, which reduce NO_x emissions.

U.S. 4,313,848 teaches countercurrent regeneration of spent FCC catalyst, without backmixing, to minimize NO_x emissions.

While such process modifications may be useful for new construction they can not be easily added to existing units, especially not to compact regenerator/stripper designs such as the Kellogg unit described above.

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U.S. 4,309,309 teaches the addition of a vaporizable fuel to the upper portion of a FCC regenerator to minimize NO_x emissions. Oxides of nitrogen formed in the lower portion of the regenerator are reduced in the reducing atmosphere generated by burning fuel in the upper portion of the regenerator.

U.S. 4,235,704 suggests that too much CO combustion promoter causes NO_x formation, and calls for monitoring the NO_x content of the flue gases, and adjusting the concentration of CO combustion promoter in the regenerator based on the amount of NO_x in the flue gas.

The approach taken in U.S. 4,542,114 is to minimize the volume of flue gas by using oxygen rather than air in the FCC regenerator, with consequent reduction in the amount of flue gas produced.

The reduction in NO_x emissions achieved by the above approaches is helpful but still may fail to meet the increasingly stringent NO_x emission limits set by local governing bodies. Much of the NO_x formed is not the result of combustion of N_2 within the FCC regenerator, but rather combustion of nitrogen-containing compounds in the coke entering the FCC regenerator.

Unfortunately, the trend to heavier feeds usually means that the amount of nitrogen compounds on the coke will increase so NO_x emissions will increase. Higher regenerator temperatures also tend to increase NO_x emissions. It would be beneficial, in existing refineries, to have a way to reduce NO_x emissions so that heavier feeds, and environmental concerns, can be accommodated.

According to the invention, it has now been realized that a better catalyst stripper design is needed to obviate the problems outlined above. Thus a better stripper would attack most of the problems in the regenerator at their source, namely poor stripping.

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Better stripping would permit increased recovery of valuable, strippable hydrocarbons and remove more hydrogen from spent catalyst to minimize hydrothermal degradation in the regenerator. It would also remove more sulfur-containing compounds from spent catalyst prior to regeneration to minimize SO_x in the regenerator flue gas and would help reduce regenerator temperature by reducing the amount of material burned in it. The present invention seeks to provide such an improved stripper, which at the same time can be incorporated into existing FCC regenerators, especially compact designs, such as the Kellog unit described above.

Accordingly, the present invention resides in a fluidized catalytic cracking process wherein a heavy hydrocarbon feed comprising hydrocarbons having a boiling point above about 343°C (650°F) is catalytically cracked to lighter products comprising the steps of:

catalytically cracking said feed in a riser reactor by mixing the feed in the base of the reactor with a source of hot regenerated catalytic cracking catalyst withdrawn from a catalyst regenerator, and cracking said feed in said riser reactor to produce catalytically cracked products and spent catalyst which are discharged from the top of the riser into a catalyst disengaging zone;

separating cracked products from spent catalyst in said catalyst disengaging zone to produce a cracked product vapor phase which is recovered as a product and a spent catalyst phase which is discharged from said disengaging zone into a catalyst stripper contiguous with and beneath said disengaging zone;

stripping said spent catalyst with steam in said stripper to produce a stripper vapor comprising cracked products and stripped catalyst;

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discharging said stripped catalyst into a catalyst regeneration zone contiguous with and beneath said stripper;

regenerating said stripped catalyst in said regeneration zone, which comprises a single dense phase bubbling fluidized bed of catalyst to which an oxygen containing regeneration gas is added and from which hot regenerated catalyst, having a temperature above that of said stripped catalyst, is withdrawn and recycled to said riser reactor, characterized in that:

said catalyst stripper is in indirect heat exchange relationship with the regeneration zone so that, during said stripping step, the spent catalyst is heated through indirect heat exchange by the hot regenerated catalyst.

The invention will now be more particularly described with reference to the accompanying drawings, in which:

Figure 1 is a schematic view of a conventional fluidized catalytic cracking unit,

Figure 2 is a schematic view of a preferred embodiment of the invention, showing a stripper heated with catalyst from a regenerator cyclone,

Figure 3 is a schematic view of a multi-stage hot stripper of the invention, with a fast fluidized bed coke combustor added to the regenerator,

Figure 4 is a schematic view of a multi-stage hot stripper of the invention, heated with catalyst from a regenerator cyclone, with a fast fluidized bed coke combustor, and

Figure 5 is a schematic view of a multi-stage hot stripper of the invention, with a preferred method of indirectly heat exchanging the catalyst in the stripper, and removing stripper vapors.

Referring to the drawings, Figure 1 is a simplified schematic view of an FCC unit of the prior art, similar to the Kellogg Ultra Orthoflow converter

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Model F shown as Fig. 17 of Fluid Catalytic Cracking Report, in the January 8, 1990 edition of Oil & Gas Journal.

A heavy feed such as a vacuum gas oil is added to the base of the riser reactor 6 via feed injection nozzles 2. The cracking reaction is completed in the riser reactor and spent catalyst and cracked products are discharged by way of 90° elbow 10 to riser cyclones 12. The cyclones 12 separate most of the spent catalyst from cracked product, with the latter being discharged into disengager 14, and eventually removed via upper cyclones 16 and conduit 18 to a fractionator (not shown).

Spent catalyst is discharged from a dipleg of riser cyclones 12 down into catalyst stripper 8, where one, or preferably 2 or more, stages of steam stripping occur, with stripping steam admitted by means not shown in Figure 1. The stripped hydrocarbons, and stripping steam, pass into disengager 14 and are removed with cracked products after passage through upper cyclones 16.

Stripped catalyst is discharged down via spent catalyst standpipe 26 into catalyst regenerator 24, with the flow of catalyst is controlled by a spent catalyst plug valve 36.

Catalyst is regenerated in regenerator 24 by contact with air, added via air lines and an air grid distributor (not shown). A catalyst cooler 28 is provided so that heat may be removed from the regenerator, if desired. Regenerated catalyst is withdrawn from the regenerator via regenerated catalyst plug valve assembly 30 and fed via lateral 32 into the base of the riser reactor 6 to contact and crack fresh feed injected via injectors 2, as previously discussed. Flue gas, and some entrained catalyst, are discharged into a dilute phase region in the upper portion of regenerator 24. Entrained catalyst is separated from

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flue gas in multiple stages of cyclones 4, and discharged via outlets 8 into plenum 20 for discharge to a flare via line 22.

Referring to Figure 2 only the differences from the unit in Figure 1 are shown and like elements in Figure 1 and 2 have like numerals. In the Figure 2 embodiment, a multi-stage hot stripper 108 is added to the lower end of the existing catalyst stripper 8 so as to extend within the regenerator, and preferably into the regenerator bubbling dense bed, so that the spent catalyst in the stripper 108 is heated by indirect heat exchange. If desired, the efficiency of the indirect heat exchange between the regenerator and the hot stripper 108 can be increased by modifying the internal or external surface of hot stripper vessel with fins, dimples, ridges or the like.

In addition, catalyst from a regenerator cyclone 118, discharged via cyclone dipleg 120 heats the catalyst from the primary stripping zone 8 by direct contact heat exchange. Stripping steam, or other stripping medium, is added by conventional steam addition means 122 and 126. The stripped hydrocarbons and stripping gas are removed either via side withdrawal means 124, preferably at multiple elevations within the stripper, or via one or more central funnels 130, defining multiple annular openings 132 connected through a central vapor outlet 134 to the vapor space above the existing stripper 8.

The embodiment shown in Figure 2 has several important advantages. A significant amount of heating of the hot stripper is effected by indirect heat exchange, because the hot stripper is within the regenerator, and preferably is at least partly immersed in the regenerator bubbling dense bed. This is beneficial in that catalyst traffic in the stripper and in the regenerator can be reduced. This reduces the size and cost of equipment and slightly reduces

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catalyst fines or dust lost from the regenerator. The hot stripper operation is improved, because the concentration of spent catalyst is higher. In contrast, relying on hot regenerated catalyst to heat the spent catalyst not only dilutes the spent catalyst, making it harder to remove the last traces of strippable hydrocarbon, but also provides additional interstitial space, and to a lesser extent, pore volume, which can entrain stripped hydrocarbons back into the regenerator.

The Figure 2 design also allows all stripped product from all stages of stripping to be withdrawn together, and sent to product fractionation. In addition, this design permits hot regenerated catalyst to be added to the hot stripper safely and in a manner which is very tolerant of failure of the hot catalyst recycle line. The latter is an important consideration for any unit which is expected to last for years in the erosive environment of a regenerator.

Figure 3 shows a preferred embodiment of the invention, with a hot stripper and a fast fluidized bed region created in the bubbling bed in the base of the regenerator. The Figure 3 embodiment employs external control of hot regenerated catalyst flow to the hot stripper.

Thus, referring to Figure 3, the stripper 8 discharges stripped catalyst into hot stripper 208 which is at least partially immersed in a high efficiency regenerator, comprising coke combustor 250 and dilute phase transport riser 252. In the coke combustor, the air admission rate, and the cross-sectional area available for flow, and catalyst addition and catalyst recycle, if any, are adjusted to maintain much or all of the bed in a "fast fluidized condition", characterized by intense agitation, relatively small bubbles, and rapid coke combustion. In terms of superficial vapor velocity and typical FCC

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catalyst sizes, this means the vapor velocity should exceed 1 m/second (3.5 feet/second), and preferably should be 1.2-4.6 m/second (4-15 feet/second), and most preferably 1.2-3 m/second (4-10 feet/second). The catalyst density in a majority of the volume of the coke combustor will be less than 560 kg/m^3 (35 pounds/cubic foot), preferably less than 480 kg/m^3 (30 pounds/cubic foot), and ideally about 400 kg/m^3 (25 pounds/cubic foot), and even less in the upper regions of the coke combustor, where the diameter of the vessel decreases.

The densities and superficial vapor velocities discussed herein presume that the unit operates at a pressure where the vast majority of FCC units operate, namely 270-380 kPa (25-40 psig). Changes in pressure change the superficial vapor velocity needed to maintain a fast fluidized bed or a bubbling dense bed. However, it is easy to calculate the superficial vapor velocity needed to support a given type of fluidization, and the bed density expected at those conditions. In general, an increase in pressure will decrease the superficial vapor velocity needed to achieve a fast fluidized bed.

The arrangement shown in Figure 3 provides a significant amount of indirect, counter-current heat exchange of spent catalyst with regenerating catalyst. The first stage of catalyst regeneration takes place in coke combustor 250, which is operates as a fast fluidized bed. Partially regenerated catalyst, and flue gas, are discharged from the fast fluidized bed region and pass as a dilute phase up transport riser 252, which encompasses the lower portion of hot stripper 208. Partially or totally regenerated catalyst and flue gas are discharged from the transport riser via cap or deflector 260, which directs catalyst and flue gas down to the bubbling dense bed. The catalyst tends to continue in a straight line to the

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bubbling dense bed, while the gas flows sideways, so a measure of catalyst separation is achieved. Catalyst discharged from cap 260 is collected as a bubbling dense bed 265. Additional regeneration gas may be added to dense bed 265, for fluffing, and preferably to obtain an additional stage of regeneration.

Because the base of the stripper 208 is well below the level of bubbling dense bed 265 it is possible to transfer catalyst from bed 265 into the hot stripper via line 220 and slide valve 222. Because of the extent of immersion of the hot stripper in the coke combustor and transport riser, and because of the intense fluidization which occurs in both of these regions, the rate of heat transfer into the hot stripper via indirect heat exchange can be very high, so relatively low rates of catalyst recycle via line 220 may be needed. This design will work well even when no catalyst is recycled, provided that conductive, rather than insulating, refractory materials are used to line the inside and outside of hot stripper 208.

Although not shown, it is possible, and usually preferred, to provide for recycling some hot regenerated catalyst from bed 265 into the coke combustor 250. If hot stripping is vigorous enough, or at least achieves a high degree of heating coupled with modest additional stripping, then catalyst recycle from the bubbling dense bed 265 to the coke combustor 250 may be greatly reduced or eliminated.

In a preferred embodiment of the invention shown in Figure 3, a significant amount of combustion air is added to bed 265 both to maintain fluidization and achieve a significant amount of coke combustion. Preferably from 5 to 60 % of the coke combustion occurs in the bubbling bed, and most preferably from 10 to 40 %. Although bed 265 is a typical fluidized bubbling bed, characterized by relatively large stagnant regions, and large bubbles of combustion air which

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bypass the bed, it is an excellent place to achieve some additional coke combustion. One of the most significant benefits of coke combustion in bubbling bed 265 is the relatively drier atmosphere. There is a lower steam partial pressure in the dense bed 265 than in a conventional dense bed regenerator, such as that shown in Fig. 1. Much of the reduction in steam partial pressure is due to the removal of water of combustion, and entrained stripping steam, with the flue gas discharged from the coke combustor. By using a flue gas/catalyst separation means on the transport riser outlet, the relatively high steam content flue gas is separated from the catalyst which is discharged down to form the bubbling fluidized bed. It is also possible to reduce the load on the cyclones above the bubbling dense bed, because much less combustion air, and consequently less entrainment of catalyst into the dilute phase, is needed when only a fraction of the coke combustion occurs in the bubbling dense bed.

Figure 4 shows another preferred embodiment with a hot stripper and a fast fluidized bed region in the bubbling bed in the base of the regenerator. The Figure 4 embodiment employs internal flow of hot regenerated catalyst to the hot stripper, as opposed to the external flow arrangement of Figure 3.

Conventional stripper 8 discharges stripped catalyst into hot stripper 308, which again extends into the regenerator. Catalyst for direct contact heat exchange of spent catalyst in the hot stripper 308 is obtained from a cyclone, preferably a primary cyclone 318, which discharges recovered, hot, regenerated catalyst via dipleg 320 into seal pot 324. This pot is designed to allow a predetermined amount of hot regenerated catalyst to flow via line 328 into the hot stripper 308, while allowing any excess material to simply overflow seal pot 324.

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This embodiment shows several preferred methods of controlling the amount of hot regenerated catalyst that is allowed to enter hot stripper 308. A plurality of steam lines, 330 and 332, are at different elevations in the hot stripper, under or near the outlet of line 328. When large amounts of fluidizing steam are added via line 330, the density of the material above line 330 is greatly reduced, which allows more catalyst to flow into the hot stripper. Addition of more steam via line 332 also reduces density, and can be used to provide better mixing of spent and regenerated catalyst, as well as provide more stripping steam.

Combustion air for the coke combustor is provided via line 301 and air ring 303. Air for additional catalyst regeneration, and for fluffing, for the bubbling dense bed is provided via air line 305 and pipe grid 307.

Fig. 5 shows a multi-stage hot stripper 508, which is intended to be mounted below a conventional steam stripper. The hot stripper 508 is at least partially, and preferably totally, immersed in the dilute phase and perhaps even the dense phase region of the bubbling bed regenerator. The stripper 508 includes a plurality of tubes 510, with inlets 512 at the top for spent catalyst, and for the discharge of stripper vapor, and outlets at the base of the tubes for discharge of stripped catalyst into the regenerator. Stripping gas, preferably steam, is admitted to a lower portion of each tube 510 to strip and aerate the spent catalyst in the tubes. The aeration should be sufficient to promote vigorous stripping, but not sufficient to blow more than minor amounts of stripped catalyst out the tops of the tubes 512.

Using multiple tubes greatly increases the surface area available for heat transfer. In a typical FCC bubbling bed regenerator, use of 20 tubes, each 2.5 cm in diameter, would increase the heating of the stripper

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that can be achieved by almost an order of magnitude as compared to the amount of stripper heating that is inherent in the Figure 1 design.

The removal of stripper vapor from the tubes is facilitated by the use of channel 518 above the tubes 510. This channel isolates the tubes from the rush of catalyst discharged by the primary stripper, and provides for a more orderly addition of spent catalyst to the stripping tubes, and for more orderly withdrawal of stripper vapor. When the tubes 510 are radially disposed, as is preferred, channel 518 may be a generally ring shaped baffle above the tubes. The hot stripper 508 should be sealed from the regenerator, which is accomplished by providing seal plate 542 around tubes 510. The tubes 510 should be made of stainless steel, or some other equivalent material which is both strong and conductive.

In an alternative embodiment (not shown) to that of Figure 5, the tubes are part of a tube and shell heat exchanger and the spent catalyst is passed through the shell side of the heat exchanger, while the tube side is open to the dilute phase region of the regenerator.

In a further embodiment (not shown), indirect heat transfer between the regenerator and the hot stripper is effected through heat pipes containing a working fluid, such as sodium, which is vaporised in the regenerator and condensed in the stripper.

FCC FEED

Any conventional FCC feed can be used. The process of the present invention is especially useful for processing difficult charge stocks, those with high levels of CCR material, exceeding 2, 3, 5 and even 10 wt % CCR.

The feeds may range from the typical, such as petroleum distillates or residual stocks, either virgin or partially refined, to the atypical, such as coal

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oils and shale oils. The feed frequently will contain recycled hydrocarbons, such as light and heavy cycle oils which have already been subjected to cracking.

Preferred feeds are gas oils, vacuum gas oils, atmospheric resids, and vacuum resids, and mixtures thereof. The present invention is very useful with heavy feeds having a metals contamination problem. With these feeds, the possibility of reduced burning load in the regenerator, and even more importantly, the possibility of a dryer regenerator, because of reduced hydrogen content of coke, will be a significant benefit.

FCC CATALYST

~~Any commercially available FCC catalyst may be~~ used. The catalyst can be 100% amorphous, but preferably includes some zeolite in a porous refractory matrix such as silica-alumina, clay, or the like. The zeolite is usually 5-40 wt.% of the catalyst, with the rest being matrix. Conventional zeolites include X and Y zeolites, with ultra stable, or relatively high silica Y zeolites being preferred. Dealuminized Y (DEAL Y) and ultrahydrophobic Y (UHP Y) zeolites may be used. The zeolites may be stabilized with Rare Earths, e.g., 0.1 to 10 Wt % RE.

Relatively high silica zeolite containing catalysts are preferred for use in the present invention. They withstand the high temperatures usually associated with complete combustion of CO to CO₂ within the FCC regenerator.

The catalyst inventory may also contain one or more additives, either present as separate additive particles, or mixed in with each particle of the cracking catalyst. Additives can be added to enhance octane (shape selective zeolites, i.e., those having a Constraint Index of 1-12, preferably ZSM-5), adsorb SOX (alumina), remove Ni and V (Mg and Ca oxides).

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FCC REACTOR CONDITIONS

Conventional riser cracking conditions may be used. Typical riser cracking reaction conditions include catalyst/oil ratios of 0.5:1 to 15:1 and preferably 3:1 to 8:1, and a catalyst contact time of 0.1 to 50 seconds, and preferably 0.5 to 5 seconds, and most preferably about 0.75 to 2 seconds, and riser top temperatures of 480 to 565°C (900 to 1050°F).

HOT STRIPPER CONDITIONS

Typical hot stripper operating conditions include temperatures which are at least 11°C (20°F) above the temperature in the conventional stripping zone, preferably at least 28°C (50°F) above the temperature in the conventional stripper, and most preferably temperatures in the hot stripper are at least 55°C (100°F) hotter.

A stripping gas or medium, preferably steam, is used for stripping. Preferably from 0.5 to 5.0 wt % steam, based on the weight of spent catalyst, is added to the hot stripping zone, in addition to the amount of stripping steam used in the conventional stripper. It is possible, and usually will optimize the overall operation of the unit, if the total amount of stripping steam used, in both the conventional stripper and the hot stripper, is roughly the same or increased only slightly. From 10 to 90 %, and preferably 20 to 60%, of the total amount of stripping medium used is added to the hot stripper. This will reduce the amount of steam added to the primary stripper, and reduce the efficiency of the primary stripper, and shift the stripping duty to the hot stripper. In this way, the overall stripping efficiency can be greatly increased, without loading up the process lines with steam, and greatly increasing the production of sour water in downstream units.

The desired heating of spent catalyst in the hot stripper can be achieved by indirect heat exchange with

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or without direct contact heat exchange (recycle of hot regenerated catalyst into the hot stripper). Each mode of heating will be briefly reviewed.

DIRECT CONTACT HEAT EXCHANGE

When direct contact heat exchange is practiced, it usually will be preferred to recycle an amount of regenerated catalyst equal to 10 to 500 % of the spent catalyst, and preferably from 15 to 150% of the spent catalyst. The heat balance equations are fairly simple, because the heat capacity of spent and regenerated catalyst is about the same. A 50/50 mix (100 % addition of regenerated to spent) of 540°C (1000°F) spent and 730°C (1350°F) regenerated catalyst will give a mix temperature of about 635°C (1175°F).

INDIRECT HEAT EXCHANGE

Conventional techniques used to calculate the amount of surface needed for heat exchange may be used. For a typical FCC regenerator, with a catalyst circulation of 15 tons per minute (13600 kg/min), it is possible to estimate fairly closely the amount of heat exchange surface needed to achieve a given temperature rise, say 56°C (100°F). The heat capacity of the FCC catalyst at these conditions is 1.17 J/kg°C (0.28 Btu/lb°F), so that 14.8×10^6 J/s (50.4×10^6 Btu/Hr) of heat must be transferred. Considering the Figure 5 embodiment, a realistic overall heat transfer coefficient is about 450 J/m²s°C (80 Btu/ft²h°F), provided that the tubes are immersed in, or are very near, a dense phase fluidized bed of catalyst. There is usually not enough heat present in most dilute phase regions to permit rapid heat transfer. One exception is the amount of heat available in a dilute phase transport riser above a coke combustor. Although this stream is, strictly speaking, a dilute phase, it is a dilute phase characterised by a very high solids content, and a high velocity. Any heat exchange tube placed in a dilute phase transport riser will exhibit an even

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higher rate of heat transfer, well in excess of the heat transfer coefficient obtainable in a classical bubbling, dense phase fluidized bed.

With a ΔT of about 167°C (300°F) (assuming 540°C for spent catalyst entering the hot stripper, 590°C for catalyst leaving the hot stripper, and a 730°C average temperature in the regenerator), about 195m^2 (2100 ft^2) will be needed. This corresponds to 20 tubes 0.3m (1 ft) in diameter and 9.1m (30 ft) long. This amount of heat exchanger surface can easily be accommodated in a conventional FCC dense bed regenerator.

When the tubes terminate in an additional vessel which is immersed in the bubbling dense bed, there is additional heat transfer into the bottom of the hot stripper via the entire "wetted" area of the immersed hot stripper

CO COMBUSTION PROMOTER

Use of a CO combustion promoter in the regenerator or combustion zone is not essential for the practice of the present invention, however, it is preferred. These materials are well-known.

U.S. 4,072,600 and U.S. 4,235,754 disclose operation of an FCC regenerator with minute quantities of a CO combustion promoter. From 0.01 to 100 ppm Pt metal or enough other metal to give the same CO oxidation, may be used with good results. Very good results are obtained with as little as 0.1 to 10 wt. ppm platinum present on the catalyst in the unit.

DISCUSSION OF HOT STRIPPING BENEFITS

The hot stripper temperature controls the amount of carbon removed from the catalyst in the hot stripper. Accordingly, the hot stripper controls the amount of carbon (and hydrogen and sulfur) remaining on the catalyst to the regenerator. This residual carbon level controls the temperature rise between the stripper and the regenerator. The hot stripper also controls the hydrogen content of the spent catalyst

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sent to the regenerator as a function of residual carbon. Thus, the hot stripper controls the temperature and amount of hydrothermal deactivation of catalyst in the regenerator.

Employing a hot stripper, to remove carbon on the catalyst, rather than a regeneration stage reduces air pollution, and allows all of the carbon made in the reaction to be burned to CO₂, if desired.

The present invention strips catalyst at a temperature higher than the riser exit temperature to separate hydrogen, as molecular hydrogen or hydrocarbons from the coke which adheres to catalyst. This minimizes catalyst steaming, or hydrothermal degradation, which typically occurs when hydrogen reacts with oxygen in the FCC regenerator to form water. The high temperature stripper (hot stripper) also removes much of the sulfur from the coked catalyst as hydrogen sulfide and mercaptans, which are easy to scrub. In contrast, burning from coked catalyst in a regenerator produces SO_x in the regenerator flue gas. The high temperature stripping recovers additional valuable hydrocarbon products to prevent burning these hydrocarbons in the regenerator.

Another benefit of hot stripping is reduced solids emissions from the regenerator. In many regenerators, solids content of flue gas is roughly proportional to the solids traffic in the dilute phase of the regenerator. Reducing the solids traffic can reduce the amount of dust and fines that escape the regenerator cyclones. In high efficiency regenerators, catalyst is recycled from a bubbling dense bed to the coke combustor, and this catalyst recycle significantly increases catalyst traffic in the regenerator. The hot stripper of the present invention allows heat to be transferred from the regenerator to the catalyst from the stripper, without recycling catalyst from the regenerator, or at least with a reduced amount of

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catalyst recirculation. This reduced catalyst load to the coke combustor reduces the amount of catalyst discharged from the coke combustor, and reduces the amount of catalyst traffic in the dilute phase region above the bubbling dense bed downstream of the coke combustor.

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Claims

1. A fluidized catalytic cracking process wherein a heavy hydrocarbon feed comprising hydrocarbons having a boiling point above about 343°C (650°F) is catalytically cracked to lighter products comprising the steps of:

catalytically cracking said feed in a riser reactor by mixing the feed in the base of the reactor with a source of hot regenerated catalytic cracking catalyst withdrawn from a catalyst regenerator, and cracking said feed in said riser reactor to produce catalytically cracked products and spent catalyst which are discharged from the top of the riser into a catalyst disengaging zone;

separating cracked products from spent catalyst in said catalyst disengaging zone to produce a cracked product vapor phase which is recovered as a product and a spent catalyst phase which is discharged from said disengaging zone into a catalyst stripper contiguous with and beneath said disengaging zone;

stripping said spent catalyst with steam in said stripper to produce a stripper vapor comprising cracked products and stripped catalyst;

discharging said stripped catalyst into a catalyst regeneration zone contiguous with and beneath said stripper;

regenerating said stripped catalyst in said regeneration zone, which comprises a single dense phase bubbling fluidized bed of catalyst to which an oxygen containing regeneration gas is added and from which hot regenerated catalyst, having a temperature above that of said stripped catalyst, is withdrawn and recycled to said riser reactor, characterized in that:

said catalyst stripper is in indirect heat exchange relationship with the regeneration zone so that, during said stripping step, the spent catalyst is

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heated through indirect heat exchange by the hot regenerated catalyst.

2. The process of claim 1 wherein the catalyst stripper is at least partially immersed in the regeneration zone.

3. The process of claim 2 wherein the catalyst stripper is at least partially immersed in the bubbling fluidized bed of the regeneration zone.

4. The process of claim 2 wherein the catalyst stripper is at least partially immersed in a coke combustor which forms part of the regeneration zone and into which spent catalyst from the stripper is discharged, the coke combustor having a base region, which is immersed in said bubbling dense bed, and an upper riser region of reduced cross sectional area relative to said base region, oxygen-containing regeneration gas being added to said coke combustor to burn coke on the spent catalyst and maintain a majority of the catalyst therein in a state of fast fluidization, the at least partially regenerated catalyst passing from said base region up through said riser region and then to bubbling dense bed.

5. The process of claim 2 wherein the catalyst stripper includes a plurality of heat exchange tubes which are at least partially immersed in the regeneration zone and which include inlets in the top portion thereof for receiving spent catalyst and outlets in the base portion thereof for discharge of stripped catalyst into the regeneration zone.

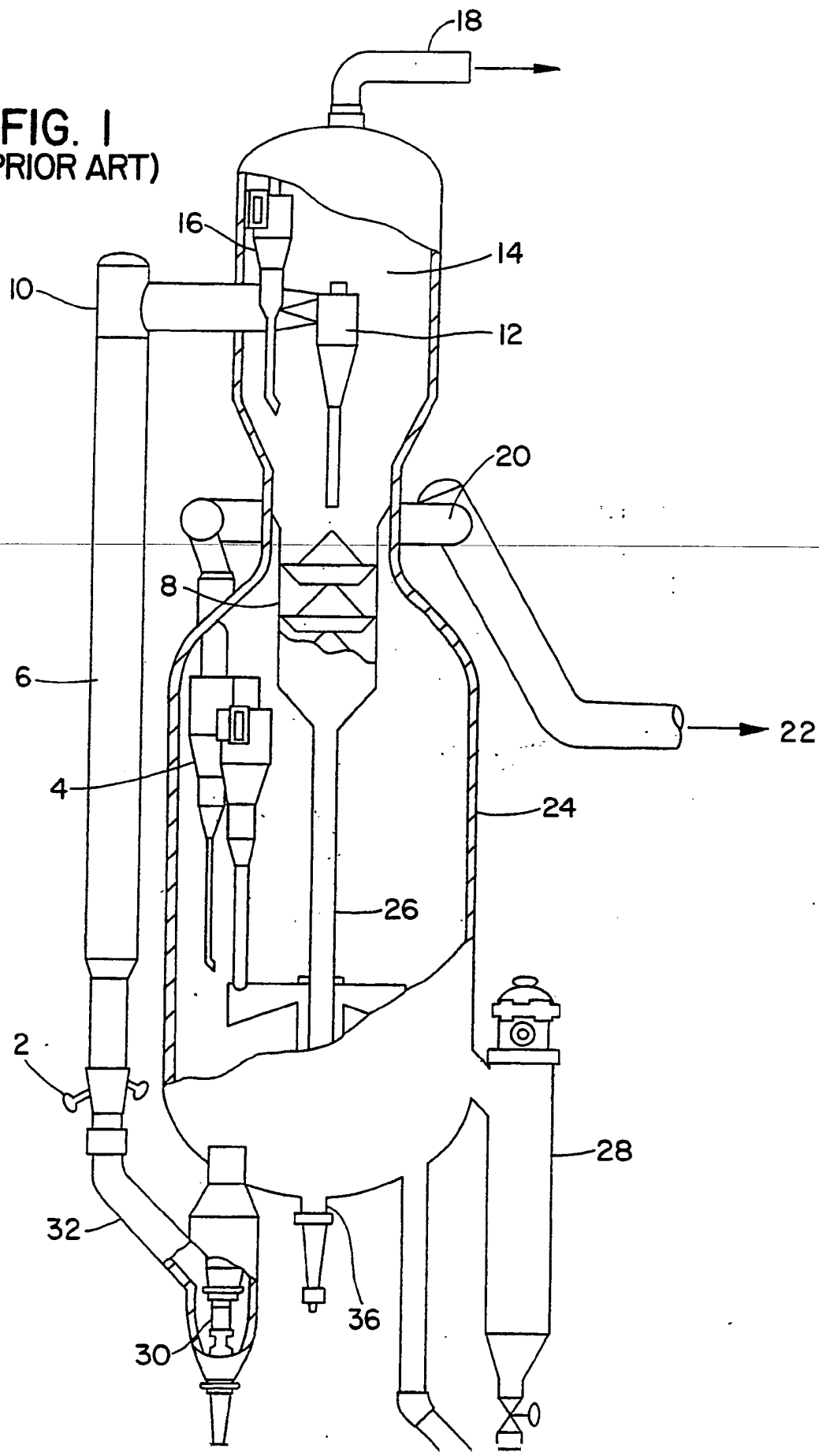
6. The process of claim 1 wherein indirect heat transfer between the regeneration zone and the catalyst stripper is effected through heat pipes extending

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therebetween and containing a working fluid which is vaporised in the regeneration zone and condensed in the stripper.

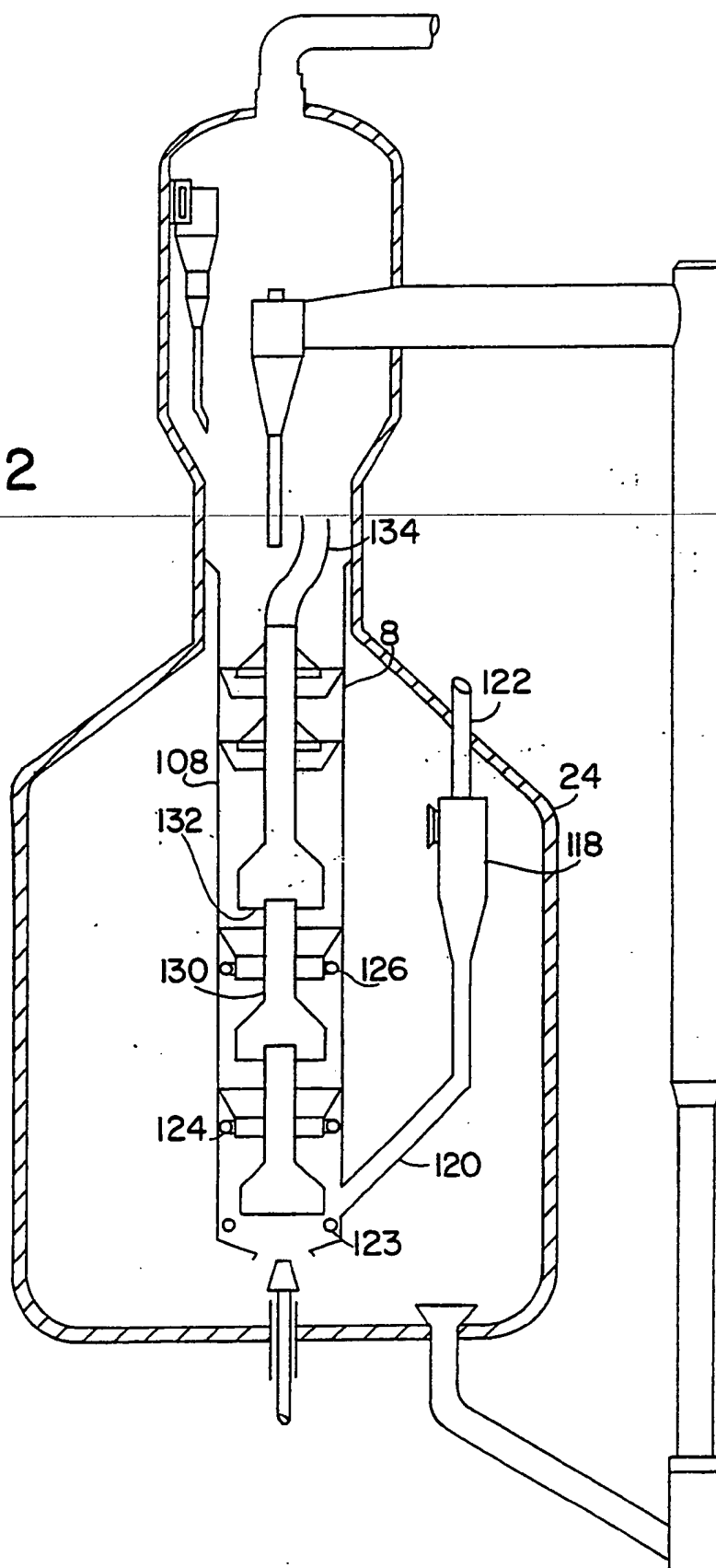
7. The process of claim 1 wherein the spent catalyst is also heated by direct contact heat exchange by regenerated catalyst recycled to the catalyst stripper from the regeneration zone.

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FIG. 1
(PRIOR ART)

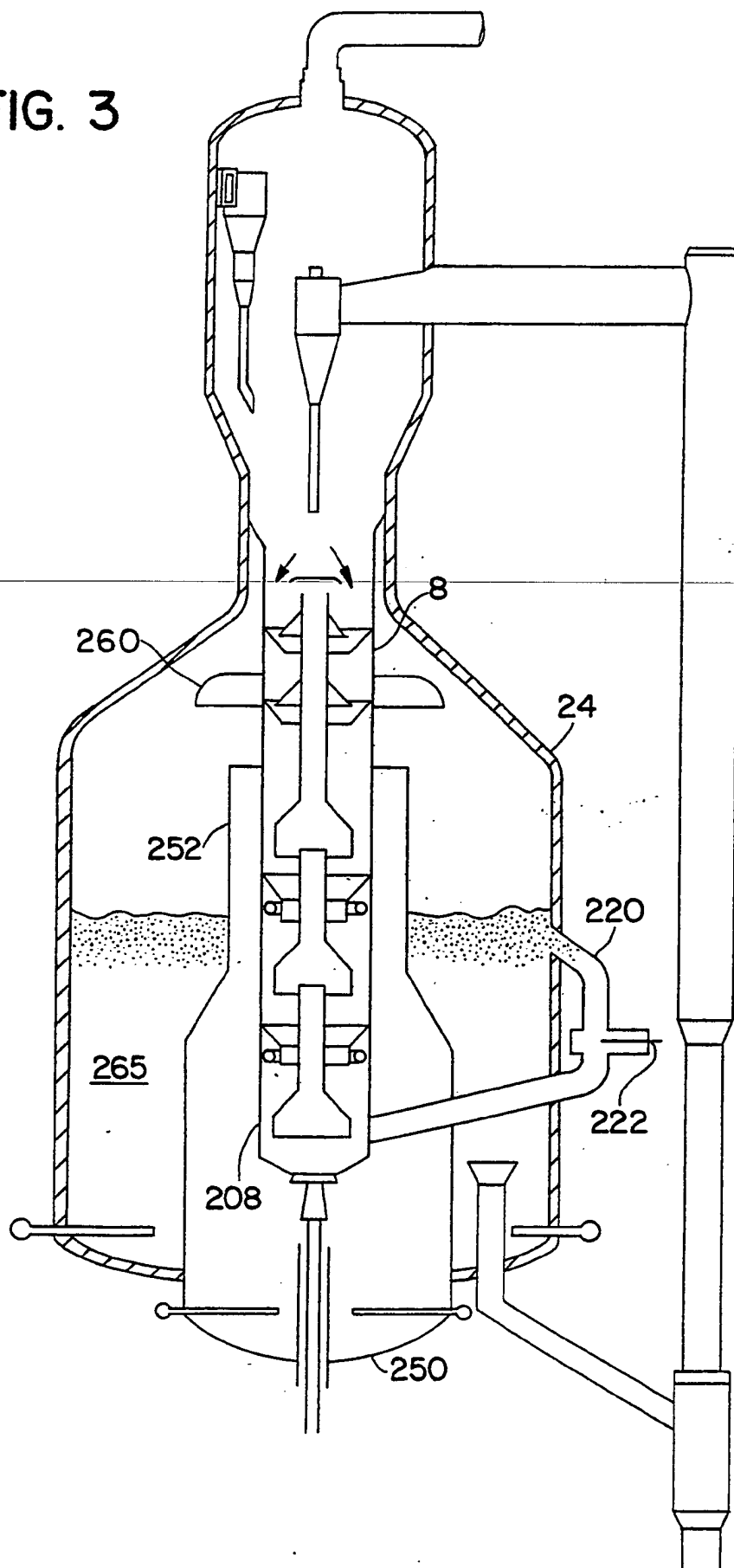
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FIG. 2



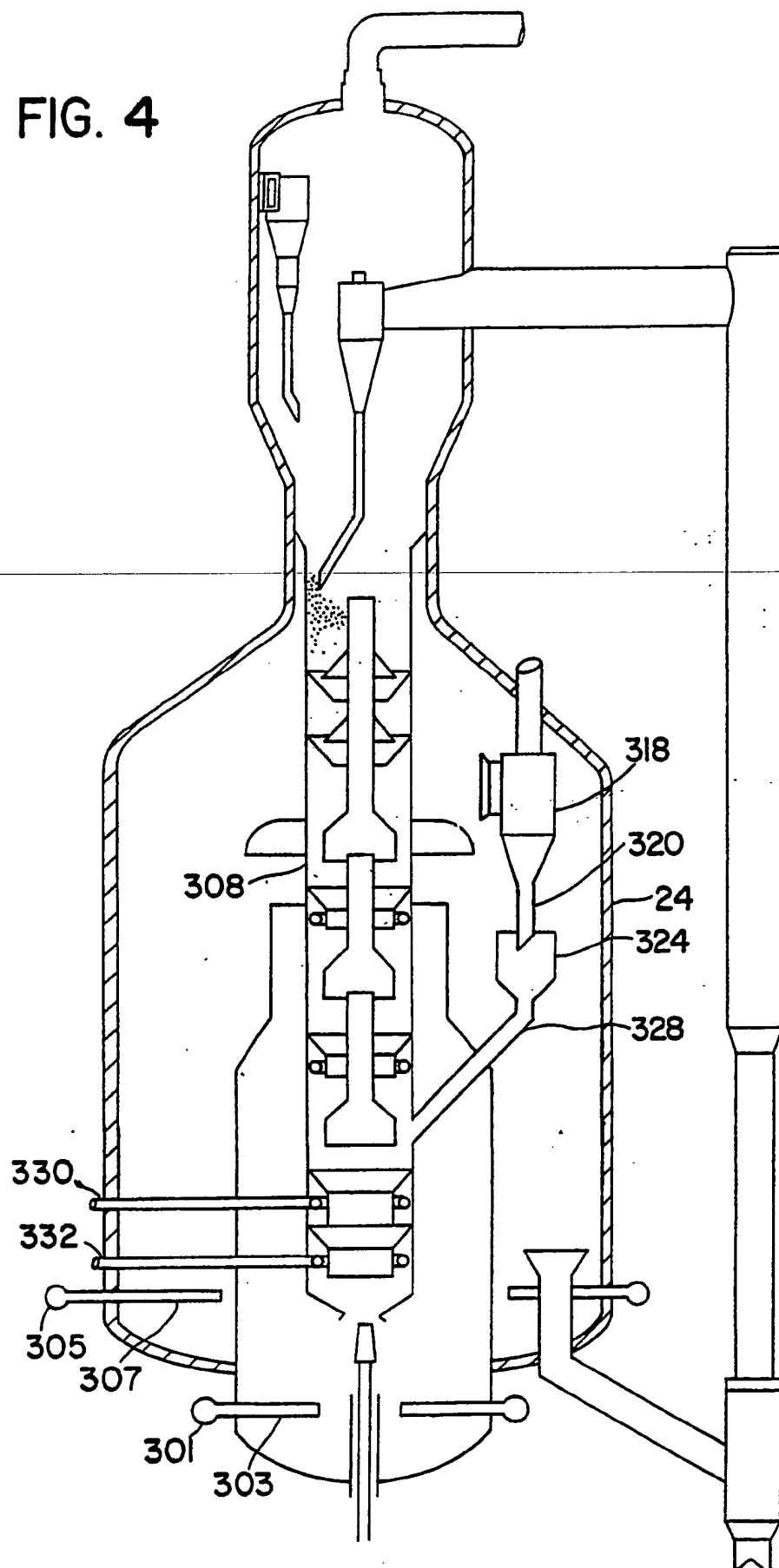
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FIG. 3

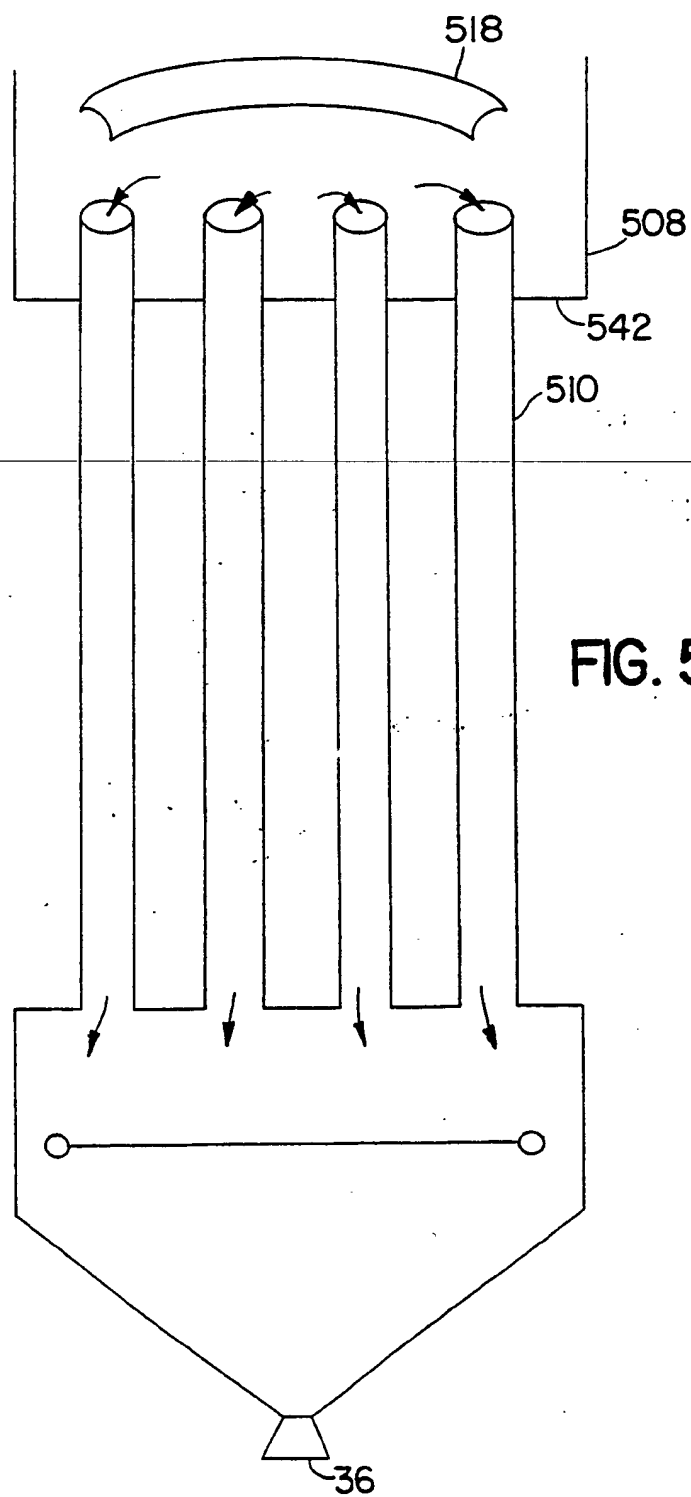


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FIG. 4



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INTERNATIONAL SEARCH REPORT

International Application No. PCT/US91/04490

I. CLASSIFICATION OF SUBJECT MATTER (if several classification symbols apply, indicate all) *

According to International Patent Classification (IPC) or to both National Classification and IPC

IPC(5): G10G 11/00; C10G 35/10

US CL : 208/113,158,159,160,164; 502/40,43,44

II. FIELDS SEARCHED

Minimum Documentation Searched *

Classification System

Classification Symbols

U.S.

208/113,158,159,160,164
502/40,43,44

Documentation Searched other than Minimum Documentation
to the extent that such Documents are included in the Fields Searched *

III. DOCUMENTS CONSIDERED TO BE RELEVANT *

Category *	Citation of Document, ** with indication, where appropriate, of the relevant passages *	Relevant to Claim No. **
X	US, A, 4,574,044 (KRUG) 04 MARCH 1986 Fig. 8 and bridging paragraph columns 13-14.	1-3
A	US, A, 4,385,985 (GROSS ET AL.) 31 MAY 1983 Fig. 1, columns 5 and 7.	4
A	US, A, 4,789,458 (HADDAD ET AL.) 06 DECEMBER 1988 Fig. 1.	7

* Special categories of cited documents: **

"A" document defining the general state of the art which is not considered to be of particular relevance

"E" earlier document but published on or after the international filing date

"L" document which may throw doubts on priority claim(s) or which is cited to establish the publication date of another citation or other special reason (as specified)

"O" document referring to an oral disclosure, use, exhibition or other means

"P" document published prior to the international filing date but later than the priority date claimed

"T" later document published after the international filing date or priority date and not in conflict with the application but cited to understand the principle or theory underlying the invention

"X" document of particular relevance: the claimed invention cannot be considered novel or cannot be considered to involve an inventive step

"Y" document of particular relevance: the claimed invention cannot be considered to involve an inventive step when the document is combined with one or more other such documents, such combination being obvious to a person skilled in the art.

"Δ" document member of the same patent family

IV. CERTIFICATION

Date of the Actual Completion of the International Search

03 SEPTEMBER 1991

International Searching Authority

ISA/US

Date of Mailing of this International Search Report

24 SEP 1991

Signature of Authorized Officer

Andrew Robinson
NHAT PHAN

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